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Study Level Design of Thorium Oxide Production

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Report on:

Study Level Design of Thorium Oxide Production

CBE 488

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1.0 Introduction

This report documents a study to provide a study-level design and analysis of a process to produce a primary product of ThO_2 (thorium oxide) and byproducts of rare earth element (REE) oxides. The given raw material is North American monazite ore with the composition below:

Composition	wt.%
La_2O_3	16.34
Ce_2O_3	33.52
ThO_2	5.40
P_2O_5	29.00
Nd_2O_3	13.75
SiO_2	1.55
U_3O_8	0.15
Total	100.00

Table 1.1: Monazite Ore Composition

The process is designed to handle an average feed rate of 1000 kg of monazite ore per hour; the economics are based on 2019 values. This process flowsheet was developed using OLI:Flowsheet. The objective of this study is to produce thorium oxide and REE oxides at purity specifications of 98% or better. In order to obtain this objective, we must produce a reasonable design that considers worker safety, environmental interactions, and long-term sustainability.

Furthermore, we have provided estimates of capital and manufacturing costs to determine the economic feasibility of the project.

The proposed process design is based on information obtained from previous studies that have explored a sulfuric acid digestion method to process the monazite feedstock and have yielded promising results. Studies presenting information useful in the development of this manuscript include: Rodliyah 2015, Amer 2013, Berry 2018, Helaly 2017, Habashi 2013, and Vijayalakshmi 2000. While all of the aforementioned articles begin with a sulfuric acid digestion, the method of precipitation of thorium varies such as utilizing pH, thermal properties, and introducing oxalic acid to the system. Other studies have suggested alternative methods or variations on the sulfuric acid digest that impact the movement of uranium in the system. The economic and safety impact of uranium in the products is a chief component to our investigation.

Monazite ore has been found in many parts of the world including India, China, and the United States. While unprocessed ore is not useful, monazite can be refined in order to produce rare earth element (REE) oxides that are used in high technology applications such as permanent magnets, optoelectronics, and superconductors (Rodliyah, 2015). Thorium-based nuclear power generation is envisioned for the future. In such a nuclear power generation, Th-232 is the fertile nuclear reactor material; when bombarded with neutrons Th-232 becomes U-233 which is a fissionable nuclear fuel. The abundance of thorium in nature along with its potential as an energy source motivates the exploration of extraction of thorium in monazite ore (World Nuclear Association, 2017).

This study-level design will illustrate and explain the flowsheet design of the process, estimate the equipment and operating costs associated with the flow sheet, assess the economic

potential of the manufacturing process, provide assumptions and background information that was used in the economic calculations, and finally discuss the feasibility for this project and recommendations going forward. This design activity is supported by the Electric Power Research Institute and the Department of Chemical and Biomolecular Engineering at the University of Tennessee, Knoxville.

2.0 Synthesis Information for Process

2.1 Overall Process Schematic

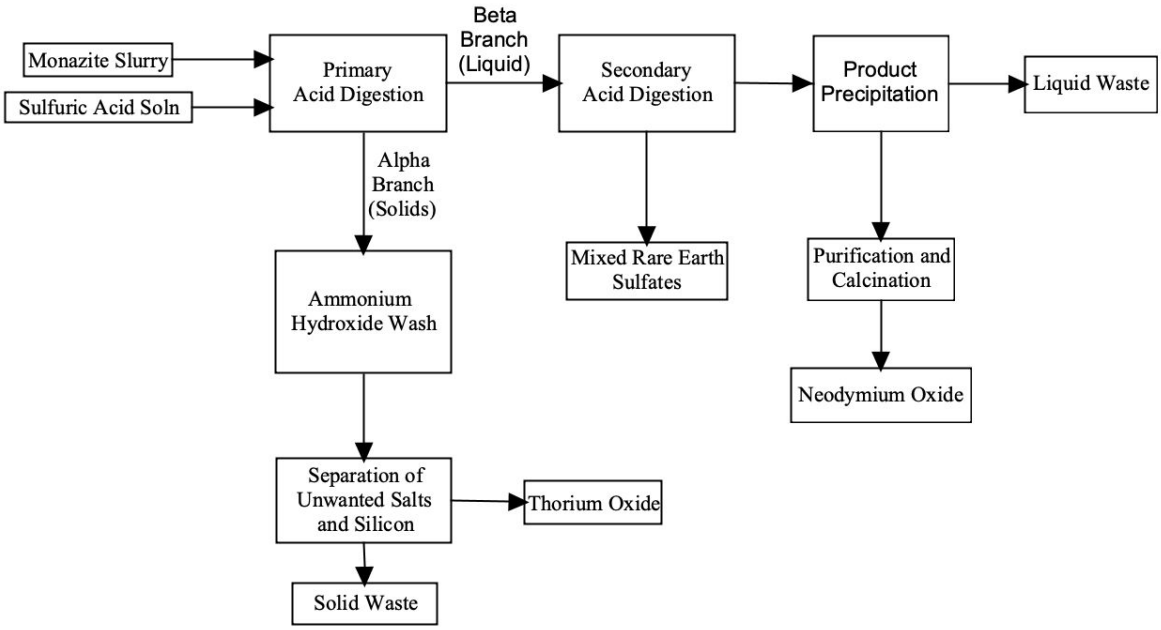

Figure 2.1: A block flow diagram depicting the proposed process.

2.2 Process Chemistry

The principal reaction for each reactor is summarized below. Note that reactions occurring reactors involved in waste treatment have been omitted for brevity as they are primarily acid-base reactions.

Initial Reactions

<u>Pre-feed Reactions: Oxides to Orthophosphates</u>

$La_2O_3 + P_2O_5 \rightarrow 2\ LaPO_4$

$Ce_2O_3 + P_2O_5 \rightarrow 2\ CePO_4$

$Nd_2O_3 + P_2O_5 \rightarrow 2\ NdPO_4$

<u>R-100 Reaction: Thorium Precipitation</u>

$ThO_2 + 2\ H_2SO_4 \rightarrow Th(SO_4)_2 + 2\ H_2O$

Alpha Branch Reaction:

<u>R-170 Reaction: Thorium Conversion</u>

$Th(SO_4)_2 + 4\ NH_4OH \rightarrow ThO_2 + 2\ (NH_4)_2SO_4 + 2\ H_2O$

Beta Branch Reactions:

<u>R-110 Reaction: Rare Earth Precipitation</u>

$2\ LaPO_4 + 3\ H_2SO_4 \rightarrow La_2(SO_4)_3 + 3\ H_2O + P_2O_5$

$2\ CePO_4 + 3H_2SO_4 \rightarrow Ce_2(SO_4)_3 + 3\ H_2O + P_2O_5$

$UO_2 + 2\ H_2SO_4 \rightarrow U(SO_4)_2 + 2\ H_2O$

<u>R-120 Reaction: Neodymium Precipitation</u>

$NdPO_4 + 2\ H_2O \rightarrow NdPO_{4\,x}\,2\ H_2O$

<u>R-130 Reaction: Neodymium Conversion</u>

$NdPO_{4\,x}\,2\ H_2O + 3\ NaOH \rightarrow Nd(OH)_3 + Na_3(PO_4) + 2\ H_2O$

<u>R-150 Reaction: Neodymium Calcination</u>

$2\ Nd(OH)_3 \rightarrow Nd_2O_3 + 3\ H_2O$

R-160 Reaction: Rare Earth Conversion

$La_2(SO_4)_3 + 2\ NaOH \rightarrow 2\ La(OH)_3 + Na_2(SO_4)$

$Ce_2(SO_4)_3 + 2\ NaOH \rightarrow 2\ Ce(OH)_3 + Na_2(SO_4)$

2.3 Literature Summary

As monazite is a good source of rare earth elements and thorium, monazite processing has been thoroughly researched for decades. Monazite has been found in a variety of regions including Egypt (Helaly, 2016) and (Saleh, 1970), Indonesia (Rodliyah, 2015), and China (Zhu, 2015). Initially, thorium was viewed as a hazardous byproduct of processing, however recent studies have shown that thorium can be utilized as a fissile fuel in place of uranium. The amount of energy that can be generated from one ton of thorium can produce the same amount of energy as two hundred tons of uranium (Helaly, 2016). The given composition of monazite shows that is abundant in rare earth elements as well as thorium. Furthermore, the sample also contains less than two percent silicon dioxide and uranium; impurities that can impact the selling price of the finished product.

While many researchers agree that strong acid or strong base digestions are the most effective methods for monazite processing, work is still being done to determine the optimal conditions to acquire the most pure products. One method of extraction is by precipitating a thorium phosphate from a monazite-sulfate solution, however there is a wide variety of poorly defined precipitate forms of thorium that can be generated. Furthermore, the gelatinous nature of this precipitate makes filtration much more complicated. With these constraints in mind, it is much simpler to precipitate thorium alongside the rare earth sulfates. This method also allows for the recycling of sulfuric acid which can reduce operating costs (Helaly, 2016).

After digestion, there are multiple paths ways to continue processing the materials. An uncommon method is using oxalate to precipitate thorium and lanthanum from the solution. This method would require further processing in order to achieve pure thorium as well as neodymium in order to achieve the maximum economic potential. Considering all of the additional separations and research this method would require, it is not likely to be a viable option for maximizing profit (Habashi, 2013).

A thorough process description for monazite processing, complete with chemistry and reaction conditions by Berry et al., detailed a process that employs a longer digestion time of four hours in the acid digestion step at a slightly higher temperature of 250 °C (Berry et al., 2018). The increased digestion time would call for a larger and more expensive reactor, though this would lead to a more pure product. In contrast, a paper by Logamtanah details a digestion step that occurs over 150 minutes at 220 °C. The shorter digestion time and lower temperature lead to a marginally less pure product, however it is still of acceptable quality, equipment costs are lower, and the process is slightly safer (Rodliyah, 2015).

With uranium often being present in monazite ore, additional separations should be incorporated in order to have pure products that are free of unwanted radioactive material. Separation of uranium in these systems has proven to be somewhat difficult. One possible method for complete extraction of uranium was proposed by Zhu et al. in which thorium and uranium are separated from rare earth elements. Uranium and thorium are then capable of being separated by utilizing N-alkyl carboxylic acid amines (R-CO-R') with varying R' groups that were described in (Preston and du Preez, 1995). While this method would separate the radioactive material from the REEs, the separation of uranium and thorium is nearly impossible

as the materials produced as mass-separating agents are uncommon and not produced commercially. Nonetheless, it may be of future significance to research production of these amines (Zhu et al., 2015).

2.4 Basic Process Economics

In addition to designing a process capable of extracting thorium oxide from monazite ore, an economic analysis will be performed to determine the viability of the proposed process. All of the equipment was designed and economic calculations were performed following the guidelines presented in Ulrich. Following these parameters, we were able to determine the net profit after taxes, raw material purchasing costs, labor and operating expenses, waste treatment, and all other miscellaneous costs associated with a chemical plant. A list of equipment and more detailed economics can be found in appendices C and D respectively.

3.0 Methods of Approach

3.1 Flow Diagram

Monazite ore is fed as a slurry into a reactor with concentrated sulfuric acid. The reactor operates at 220 °C and 15 atm in order to convert the rare earth oxides into rare earth sulfates. After the initial acid digestion, the components are cooled in a heat exchanger before being sent to a filter. In the filter, the process splits into the alpha branch which produces thorium oxide and a beta branch which produces neodymium oxide. In the alpha branch, solid silicon dioxide and thorium sulfate are separated by density in a component splitter to isolate the thorium. Thorium sulfate is then fed into a reactor alongside ammonium hydroxide in order to reconvert it back to

thorium oxide. The mixture is then fed to another component splitter which isolates the thorium oxide based on density.

In the beta branch, rare earth phosphates from the first reactor are fed to a second reactor along with more sulfuric acid and water at 275 °C and 20 atm in order to convert all of the remaining rare earth phosphates, with the exception of neodymium, to sulfates. Once the stream exits the reactor, it is cooled and filtered so that the solid rare earth sulfates and uranium sulfate are removed from the process to be sold as a byproduct. The neodymium phosphate is then fed to a reactor with sodium hydroxide solution in order to form a solid neodymium phosphate hydrate that can be filtered out from the remaining waste products. After filtration, more sodium hydroxide solution is fed to a reactor with the neodymium phosphate hydrate in order to form the neodymium hydroxide. The hydroxide is separated from the residual contaminants in the stream via a component splitter and filter before being calcined to form the neodymium oxide.

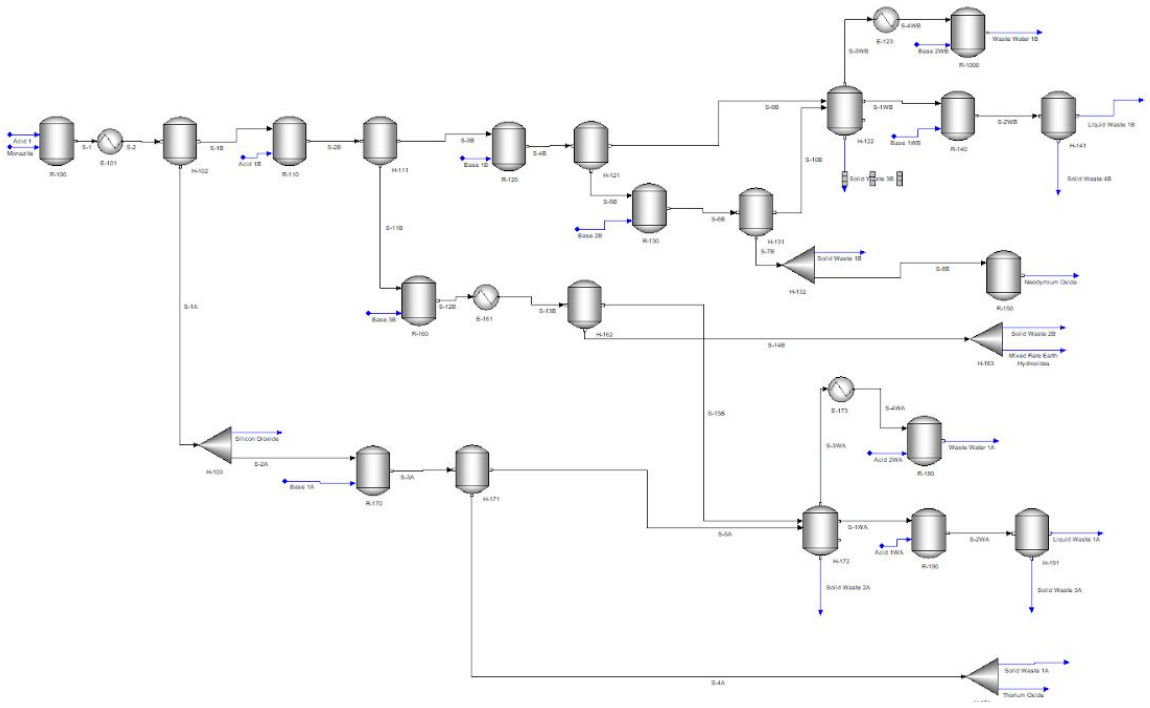

Figure 3.1: Final flowsheet of the proposed process

3.2 Sustainability, Environmental Concerns, and Worker Safety

As mentioned earlier, thorium oxide has incredible potential as a fissile fuel as it can produce approximately 200 times the amount of energy compared to the same amount of uranium. Furthermore, thorium is far more abundant in nature. With this in mind, efficient production of thorium would have a tremendous impact on the energy industry and allow us to move away from energy sources that are less efficient and environmentally friendly such as fossil fuels.

In terms of the process's sustainability, one of the major drawbacks to this proposed plant is that it consumes 4638 kilograms of water per hour. With this rate of consumption, it may be wise to consider implementing a water recycle in the system in order to reduce the amount of water used from outside sources. This is outside the scope of this study level design, however economic analysis should be conducted to determine whether the cost of the water recycling system would significantly reduce the cost of water.

Unwanted byproducts of this process include a variety of salts as well as a collection of residual rare earth elements. The salts and residual rare earth elements are to be separated from the product and disposed of. The disposal of these materials is not accounted for in this study level design, however in our analysis we have accounted for a solid waste disposal cost of $.50 per metric ton of material.

Thorium oxide should be handled and stored with extreme caution as it is a radioactive substance that can lead to multiple health issues including cancer. Furthermore, the mixed rare earth stream contains .3 mol% uranium sulfate which, while not present in high amount, still can still be considered a health hazard.

Outside of radiation concerns, many vessels in this process are operating at considerably high temperatures and pressures. Extreme caution should be taken in order to ensure that maintenance is performed regularly to prevent any potential accidents. The concerns for proper operation are heightened by the use of highly caustic acids and bases such as sulfuric acid and ammonium hydroxide throughout the entire process. With the amount of acid being used in the system, it is critical to ensure that all product and waste streams are properly neutralized before being removed from the process.

3.3 Product Purity

The primary product, thorium oxide, can be collected at a purity of ~98% or higher. Neodymium oxide, the secondary product, is also collected at a purity of ~98% or higher. A tertiary product, the mixed rare earth hydroxide, has the following mass composition: 66.36 % cerium hydroxide, 33.38% lanthanum hydroxide and .26% uranium oxide.

4.0 Results

After designing a process flow diagram in OLI, an economic analysis was performed to determine the viability of the process. While thorium oxide is the primary product, it is not the most profitable product. Though neodymium oxide has a lower selling price than thorium oxide, it is nearly three times more abundant in the given composition of monazite ore. The figure below summarizes the percentage of the annual sales income that can be attributed to each product. Note that because the lanthanum and cerium products are mixed and contain a small amount of uranium, they are sold at half their listed prices. Given the presented profit

distribution, it is critical that there be a demand for neodymium oxide to ensure maximum profit. Before expenses, the gross profit from sales is determined to be $86,686,156.80.

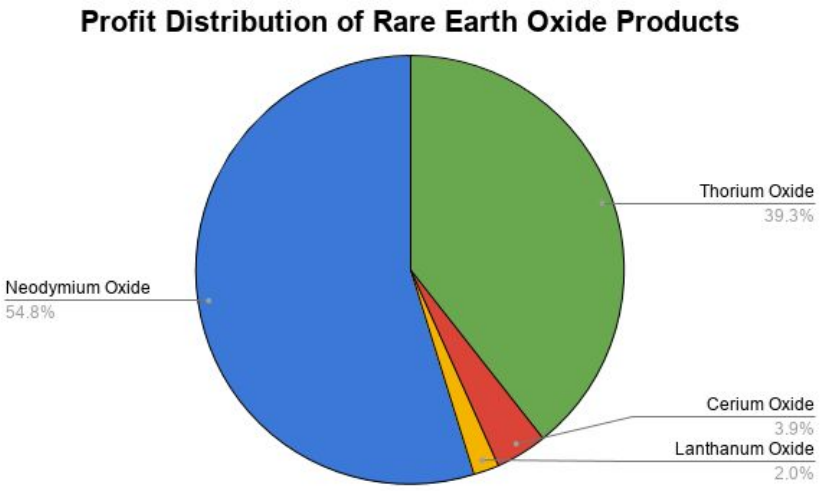

Figure 4.1: Distribution of revenue among products from the proposed process

4.1 Capital Cost Estimates

Equipment was designed based on volumetric flow requirements. All equations and values were taken from Ulrich. Some generalizations were made as Ulrich does not provide information on sizing cyclone separators. Sample calculations for each equipment type can be found in Appendix D and individual equipment costs can be found alongside the capital cost summary in Appendix C. The total grass-roots capital investment of this plant was determined to be $36,064,510.65.

4.2 Manufacturing Costs

The yearly manufacturing results were also calculated using the methods presented in Ulrich. A full table detailing the manufacturing costs can be found in Appendix C. Annual

manufacturing expenses were calculated to be $26,994,479.08. A distribution of manufacturing expenses are presented in the table below:

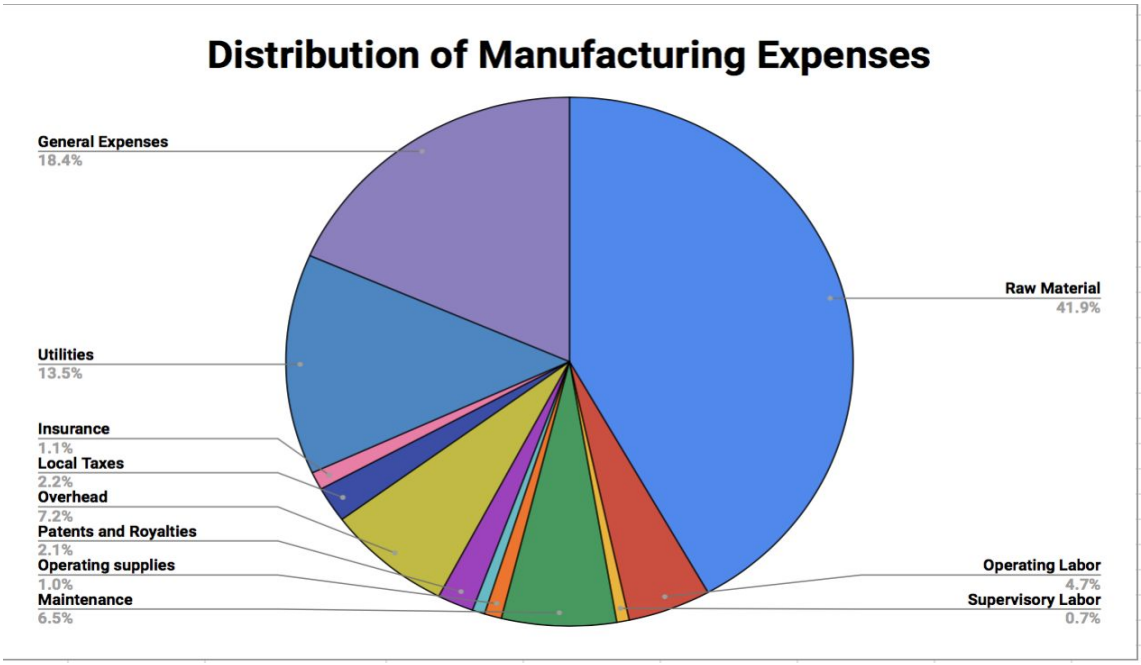

Figure 4.2: A distribution of manufacturing expenses in the proposed process

5.0 Discussion of Results

Our study shows that this process is a viable option for extracting thorium oxide as well as other REE oxides from monazite ore at a high purity. After manufacturing expenses, depreciation, and income taxes, the annual net profit was determined to be $30,145,950.72. Assuming a three year construction period, a cash flow analysis was conducted on the process to evaluate the project using undiscounted and discounted (10% and 15%) cash flows. It is determined that the plant would begin to break even approximately 4.75 - 5.5 years after construction begins; details can be found in Appendix C. Though the initial cost of grass-roots

capital is somewhat high, the plant has a short payback period due to the high economic potential of the plant.

Raw materials account for approximately 42% of the annual manufacturing expenses; of the raw material costs, approximately 35% comes from purchasing reagents such as acids, bases, and process water. With this in mind, there may be further optimization that can be done in order to reduce the total cost of raw materials.

When designing liquid cyclone separators, they typically only effective in processes that contain materials with a minimum particle diameter of 10 μm. The material in the proposed process streams are predominantly sulfates, we estimated this material to have a particle diameter resembling calcium sulfate (7-9μm according to usg.com). Since this is close to the limit of the unit effectiveness, it would be beneficial to consult with a manufacturer to ensure the units operate effectively. Further research and development may need to be done on the part of the manufacturer to design units capable of processing particles with a diameter of 7 μm.

6.0 Conclusions

Considering the calculated economic potential of the proposed process, this is a viable option for extracting thorium oxide, neodymium oxide, and mixed REE hydroxides. While neodymium oxide does not have the highest selling price, it is nearly three times more abundant than thorium oxide in the given monazite composition and therefore has the highest economic potential of any product. As long as the selling price of neodymium oxide does not significantly drop the plant should remain profitable; earning nearly $30,000,000 per year after taxes.

Furthermore, the proposed process utilizes floatation methods to separate components based on density. While the separated components have a great difference in their respective densities, it is unlikely that the separation will be perfect. For this reason, the purity of these products is described as ~98% even though OLI calculated perfect separations in these processes.

7.0 Recommendations for Optimization

After performing an economic analysis on the proposed process, we find that this process would be a viable option for processing monazite ore. In order to ensure the process remains profitable, it is imperative that there is a demand for neodymium oxide as it accounts for the majority of profits from sales. While the proposed process is a viable option, there are various areas in which modifications could be introduced to improve plant safety and to reduce overall costs.

In terms of purchasing materials, this plant consumes 4638 kilograms of water per hour; it may be beneficial to incorporate a water recycle system into the design in order to reduce utility costs. Aside from water, strong acids and bases are also used liberally in this process. To circumvent the cost of reagents, it may be beneficial to explore adding an element such as an electrodialysis unit to convert the reacted ion salts back into their respective acids and bases.

Much research was conducted at the beginning of this project to make the operating conditions safer for the plant employees. From the existing literature, we found that it was nearly impossible to carry out these reactions and separations without the extreme pressures, temperatures and caustic materials. Nevertheless, if research funding is available, more extensive research should be performed with the goal of improving process safety in the plant.

8.0 References

Berry, Loren. “Canadian Metallurgical Quarterly.” *Taylor and Francis Online*, 30 May 2018, www.tandfonline.com/doi/pdf/10.1080/00084433.2018.1478490.

Habashi, F. “Extractive Metallurgy of Rare Earths.” *Taylor & Francis*, Canadian Metallurgical Quarterly, 18 Nov. 2013, www.tandfonline.com/doi/abs/10.1179/1879139513Y.0000000081.

Helaly, O.S. “Application of the Alternative Traditional and Selective Precipitation Routes for Recovery of High Grade Thorium Concentrates from Egyptian Crude Monazite Sand.” *Arab Journal of Nuclear Sciences and Applications*, vol. 50, 5 Dec. 2016.

“High-purity additives for high-quality food, beverage and pharmaceutical products.” *United States Gypsum Company*, https://www.usg.com/content/dam/USG_Marketing_Communications/united_states/product_promotional_materials/finished_assets/calcium-sulfate-filler-brochure-en-IG172.pdf. Accessed 4 May 2019.

“High Temperature Properties.” *Stainless Steel Information Center*, http://www.ssina.com/composition/temperature.html. Accessed 4 May 2019.

Preston, John S, and Anna C du Preez. “SOLVENT EXTRACTION OF URANIUM(VI) AND THORIUM(IV) FROM NITRATE MEDIA BY CARBOXYLIC ACID AMIDES.” *Taylor & Francis*, 21 June 2010, www.tandfonline.com/doi/abs/10.1080/07366299508918282.

Rodliyah, Isyatun, et al. “EXTRACTION OF RARE EARTH METALS FROM MONAZITE MINERAL USING ACID METHOD.” *Indonesian Mining Journal*, jurnal.tekmira.esdm.go.id/index.php/imj/article/view/305.

Saleh, F. A., and A. K. Farag. “Recovery of Uranium, Thorium and Cerium from Egyptian Monazite Sands. I.” *Zeitschrift Für Anorganische Und Allgemeine Chemie*, John Wiley & Sons, Ltd, 9 Nov. 2004, onlinelibrary.wiley.com/doi/abs/10.1002/zaac.19703750314.

“Sulphuric Acid on the Web.” *Sulphuric- Acid*, http://www.sulphuric-acid.com/techmanual/Properties/properties_acid_viscosity.htm. Accessed 4 May 2019.

“Thorium.” *Thorium - World Nuclear Association*, World Nuclear Association, Feb. 2017, www.world-nuclear.org/information-library/current-and-future-generation/thorium.aspx.

Ulrich, Gael D., and Palligarnai T. Vasudevan. *Chemical Engineering Process Design and Economics: a Practical Guide*. 2nd ed., Process Publishing, 2004.

Vijayalakshmi, R. “Processing of Xenotime Concentrate by Sulphuric Acid Digestion and Selective Thorium Precipitation for Separation of Rare Earths.” *Hydrometallurgy*, Elsevier, 8 June 2001, www.sciencedirect.com/science/article/pii/S0304386X00001596.

Zhu, Zhaowu. “Separation of Uranium and Thorium from Rare Earths for Rare Earth Production – A Review.” *Minerals Engineering*, Pergamon, 6 Apr. 2015, www.sciencedirect.com/science/article/pii/S0892687515001028.

Appendices

Appendix A: Project Assumptions

- Cyclone separators are typically designed in close conjunction with the equipment manufacturer. Because of time constraints, it was assumed that the particle size of the materials is sufficient for cyclone operation.
- Flotation works 100%
- Assumption for specific design are outlined in appendix D.

Appendix B: Stream Compositions and Mass and Energy Balances

Inlets (kg/hr)																			
Stream	Monazite	Acid-Water	Base 1A	Acid 1WA	Acid 2WA	Acid 1B	Base 1B	Base 2B	Base 3B	Base 1WB	Base 2WB								
Water	1000	663				400	500	50	2000										
Sulfuric Acid		1612	25	122.6	65.9	1500													
Sodium Hydroxide							1000	100	500	353.3	0.3044								
Ammonium Hydroxide			50																
Lanthanum Phosphate	241.99																		
Cerium Phosphate	480.15																		
Thorium Oxide	54																		
Pyrophosphate	14.44																		
Neodymium Phosphate	195.51																		
Silicon Dioxide	15.5																		
Uranium Oxide	1.443																		

Outlets (kg/hr)																			
Stream	Thorium Oxide	Mixed RE Hydroxide	Neodymium Hydroxide	Silicon Dioxide	Solid Waste 1A	Solid Waste 2A	Solid Waste 3A	Liquid Waste 1A	Waste Water 1A	Solid Waste 1B	Solid Waste 2B	Solid Waste 3B	Solid Waste 4B	Liquid Waste 1B	Waste Water 1B				
Water					Tr			0.049126905	1993.61196				Tr	0.13696756	3024.880635				
Thorium Oxide	54																		
Neodymium Hydroxide			90.35908829																
Cerium Hydroxide		390.3193484					0.013379												
Lanthanum Hydroxide		196.3572312				Tr													
Uranium Oxide		1.441187914			0.001850803	Tr		Tr											
Silicon Dioxide				15.1903268								0.160580022	0.14800287	Tr					
Ammonium Sulfate					4.035370604														
Thenardite						625.3737084	83.00774988												
Sodium Sulfite Decahydrate							215.1106418				84.02489832								
Sodium Ammonium Sulfate Tetrahydrate							1.125322888												
Ammonia								0.004350944	23.14291497										
Sulfur Trioxide		0.663675491	66.36754915					0.020440579	53.79532186										
Sodium Hydroxide		0.333873999	33.38739995					0.014010948						0.054050346	0.304597568				
Cerium Oxide		0.002450509	0.249505095				Tr												
Lanthanum Sulfate							Tr												
Sodium Orthophosphate										135.6388059									
Sodium Bisulfate												3494.951016							
Thorium Sulfate											Tr								
Cerium Orthophosphate Dihydrate													0.07294104						
Lanthanum Orthophosphate Dihydrate													0.230556852						
Sodium Pyrophosphate Decahydrate													411.9573748						
Sodium Dihydrogen Orthophosphate Dihydrate													220.6992393						
Neodymium Orthophosphate Dihydrate													4.843178088						
Lanthanum Phosphate														Tr					
Cerium Phosphate														Tr					
Neodymium Phosphate														Tr					
Pyrophosphate															0.066787454				

Table B.1a Composition of inlet and outlet streams in the proposed process

Intermediate Streams (kg/hr)																									
Stream	S-1	S-2	S-1A	S-2A	S-3A	S-4A	S-5A	S-1B	S-2B	S-3B	S-4B	S-5B	S-6B	S-7B	S-8B	S-9B	S-10B	S-11B	S-12B	S-13B	S-14B	S-15B			
Water	1907.91	1959.09			50.1523		50.1523	1959.09	2633.27	2633.27	3105.25		79.3715			3105.25	79.3715		1927.64	1927.54		1927.54			
Sulfur Trioxide	1108.44	1283.16			30.3039		30.3039	1283.16	2131.4	2131.4	2137.37					2137.37			322.518	322.905		322.905			
Uranium Oxide	Tr	1.443						1.443	Tr	Tr									1.443	1.443	1.441	Tr			
Pyrephosphate	117.187	14.44						14.44	232.748	232.748	232.748					232.748									
Thorium Oxide	0.00308	Tr			54	54	Tr	Tr	Tr	Tr	Tr					Tr									
Cerium Sulfate	411.461								580.404									580.404							
Cerium Orthophosphate Dihydrate	105.96																								
Lanthanum Orthophosphate	164.595	241.99						241.99	0.199777	0.199777	0.199777					0.199777									
Lanthanum Orthophosphate Dihydrate	89.3188																								
Neodymium Orthophosphate	28.3203	195.51						195.51	195.51	195.51	0.509578		0.002848			0.509578	0.002848								
Neodymium Orthophosphate Dihydrate	192.372										224.372	224.372													
Cerium Orthophosphate	47.928	47.928						480.15	0.063248	0.063248	0.063248					0.063248									
Silicon Dioxide	15.5	15.5	15.191					0.30861	0.30861	0.30861	0.30861					0.30861									
Thorium Sulfate	86.744	86.744	86.749	86.749					2.29872																
Uranium Sulfate	2.29872	2.29872							7.31699	7.31699								2.29872							
Sulfuric Acid				Tr			Tr																		
Ammonia					23.2574		23.2574																		
Ammonium Sulfate					4.03539	4.03539																			
Lanthanum Sulfate									292.576									292.576	Tr	Tr		Tr			
Sodium Hydroxide											1000		2.1863			1000	2.1863		452.292	452.679		452.679			
Sodium Orthophosphate													133.64	133.64											
Neodymium Hydroxide													159.172	156.154	156.154		3.01759								
Cerium Oxide																			0.081998	0.01156		0.01156			
Cerium Hydroxide																			390.241	390.323	390.323				
Lanthanum Hydroxide																			196.354	196.354	196.354				
Thenardite																			84.713	84.027	84.027				

Table B.1b Composition of intermediate streams in the proposed process

Waste Streams (kg/hr)										
Stream	S-1WA	S-2WA	S-3WA	S-4WA	S-1WB	S-2WB	S-3WB	S-4WB		
Water	75.4733	0.049129	1981.55	1981.55	127.174	0.136599	3024.86	3024.93		
Ammmonia	0.11505	0.00435	23.1423	23.1423						
Sulfur Trioxide	0.652818	0.02044	Tr		20.7094	0.0158156	0.00023378	0.304598		
Uranium Oxide	Tr	Tr								
Sodium Hydroxide	100.429	0.0140336								
Cerium Oxide	0.01156	Tr								
Lanthanum Sulfate	Tr	Tr								
Cerium Hydroxide		0.01347								
Lanthanum Hydroxide		Tr								
Thenardite		83.0098				296.34				
Sodium Sulfate Decahydrate		215.105								
Sodium Ammonium Sulfate Tetrahydrate		1.12531								
Sulfuric Acid			Tr		179.27		0.37285			
Sodium Hydroxide					18.0349	0.0540507				
Pyrophosphate					231.651	0.0667896				
Thorium Oxide				Tr	Tr					
Lanthanum Orthophosphate					0.199777	Tr				
Neodymium Orthophosphate					4.20922	Tr				
Cerium Orthophosphate					0.063248	Tr				
Silicon Dioxide					0.148019	0.148019				
Cerium Orthophosphate Dihydrate						0.0729416				
Lanthanum Orthophosphate Dihydrate						0.230555				
Sodium Pyrophosphate Decahydrate						411.954				
Sodium Dihydrogen Orthophosphate Dihydrate						220.897				
Neodymium Orthophosphate Dihydrate						4.84322				

Table B.1c Composition of waste streams in the proposed process

*Note: Tr indicates there is trace amounts of this material present. These compounds are typically present in 10^{-6} kg/hr or less.

Energy Balances

The major sources of energy exchange in our process design are attributed to the four heat exchangers and the 10 reactors. OLI determined the heat duty required for each of these units. Using this information, we could determine the amount of cooling water or steam needed in the system to have a net energy value of zero. The following equation was used extensively to find the water requirement:

$$Q_{Process} = m_{steam\ or\ cw} * \Delta h_{steam\ or\ cw} * \varepsilon. \qquad \textbf{Equation 1}$$

Where Q represents heat duty, $m_{steam\ or\ cw}$ is the mass of steam (or cooling water), $\Delta h_{steam\ or\ cw}$ is enthalpy change of the material, and ε is efficiency of the heat exchange.

E-173

Analyzing the unit in OLI, we see that the process heat duty is negative so the process material will need to lose energy to reach its desired state. In this situation, cooling water is used to absorb the heat that the process stream is losing. We assume an efficiency value of 0.97 and a cooling water temperature change of 15˚C (making sure that the cooling water it's expected outlet flow is colder than the process steam conditions), in order to find the desired mass flow rate. Thus,

$Q_{Process}$ = -1498.08 kJ/s (Obtained from OLI flowsheet)

ε = 0.97 (Assumed value based on U-tube calculations in Ulrich)

$$\Delta h_{cw} = C_p * \Delta T = 4.19\ kJ/kg^{\circ}C\ * 15^{\circ}C = 62.85\ kJ/kg$$

We can solve by rearrange equation 1 to the following form:

$$m_{cw} = \frac{Q_{Process}}{\Delta h_{cw} * \varepsilon} \qquad \textbf{Equation 2}$$

Plugging in values, we find that that:

$$m_{cw} = \frac{1{,}498.08\ kJ/s}{62.85\ kJ/kg * 0.97} = 24.57 \text{ kg/s}$$

E-123

This unit uses the same assumptions and logic as E-173 as it's heat duty is also negative . The only values that changes is $Q_{Process}$. Giving the following calculation and cooling water flow rate:

$$Q_{Process} = \text{-2,312.14 kJ/s (Obtained from OLI flowsheet)}$$
$$m_{cw} = \frac{2{,}312.14\ kJ/s}{62.85\ kJ/kg * 0.97} = 37.9 \text{ kg/s}$$

E-101

This unit also has a negative heat duty. However, the method of determining cooling water flow rate was slightly different. Instead of solving for m_{cw} explicitly, a guess and check method was implemented. In this case, we guessed a value for m_{cw} and checked to see if it could satisfy the needs of the system. The process stream for the heat exchanger has an inlet temperature of 220˚C and a desired outlet temperature of 25˚C. Assuming our cooling water has an inlet around 9˚C, we need a cooling water outlet temperature no greater than 25˚C for the operation to be feasible. Using the guess and check method, we find that assuming a 35,000 kg/hr cooling water flow rate yields a temperature below 25˚C. The following calculation is evidence:

$$Q_{Process} = \text{-1,992,086.08 kJ/hr (Obtained from OLI flowsheet)}$$
$$\varepsilon = 0.97 \text{ (Assumed value based on U-tube calculations in Ulrich)}$$
$$T_{initial} = 9^{\circ}C$$
$$C_p = 4.19\ kJ/kg^{\circ}C \text{ (Value for water)}$$

Equation 1 can be rearranged to the following form:

$$T_{final} = \frac{Q_{Process}}{m_{cw} * \varepsilon * C_p} + T_{initial} \qquad \textbf{Equation 3}$$

When plugging in values, $T_{final} = 23^{\circ}C$.

E-161

This unit uses the same methodology as E-101. The heat duty is negative, with the process stream cooling down from 100˚C to 70˚C. We assume a 10,000 kg/hr flow cooling water flow rate and find that the final temperature of the cooling water is below 70˚C making it feasible. This is shown in the following calculation:

$$Q_{Process} = \text{-291,296.77 kJ/hr (Obtained from OLI flowsheet)}$$
$$\varepsilon = 0.97 \text{ (Assumed value based on U-tube calculations in Ulrich)}$$
$$T_{initial} = 9^{\circ}C$$
$$C_p = 4.19\ kJ/kg^{\circ}C \text{ (Value for water).}$$

Using Equation 3, $T_{final} = 16.16^{\circ}C$.

Reactors

OLI calculated a heat duty that went along with each reactor during the process with the exception being the calciner. If the reactor heat duty was negative, we chose add a cooling water jacket to the vessel so it could absorb the heat released from the reaction. If the heat duty was positive, we designed a steam jacket to add heat to the vessel contents so it could stay at the optimum reaction temperature.

The following reactors have negative heat duties: R-120, R-140, R-180, R-190, and R-1000. An energy balance was conducted on these units by finding the cooling water mass flow rate in the jacket. We used the OLI values, assumed an efficiency value of 0.80, and a cooling water temperature change of 15˚C (making $\Delta h_{cw} =$ 62.85 kJ/kg) to plug into equation 2. The following values were calculated:

R-120: $Q_{Process} =$ -1,448.17 kJ/s

$$m_{cw} = \frac{1{,}448.17\ kJ/s}{62.85\ kJ/kg * 0.80} = 28.8 \text{ kg/s}$$

R-140: $Q_{Process} =$ -131.68 kJ/s

$$m_{cw} = \frac{131.68\ kJ/s}{62.85\ kJ/kg * 0.80} = 2.62 \text{ kg/s}$$

R-180: $Q_{Process} =$ -1,527.28 kJ/s

$$m_{cw} = \frac{1{,}527.28\ kJ/s}{62.85\ kJ/kg * 0.80} = 30.38 \text{ kg/s}$$

R-190: $Q_{Process} =$ -111.46 kJ/s

$$m_{cw} = \frac{111.46\ kJ/s}{62.85\ kJ/kg * 0.80} = 2.22 \text{ kg/s}$$

R-1000: $Q_{Process} =$ -0.215 kJ/s

$$m_{cw} = \frac{0.215\ kJ/s}{62.85\ kJ/kg * 0.80} = 0.0043 \text{ kg/s}$$

The remaining reactors have positive heat duties: R-100, R-110, R-130, R-160, and R-170. The reaction temperatures for these reactors range from 100˚C to 275˚C. Since heat is being lost from these units, we need a higher temperature jacket to provide a heat gradient so the temperature of the vessel contents can remain constant (as it is modeled in OLI). We chose to flow steam at a pressure of 70 bar through the reactor jacket. This was chosen due to the fact that the temperature of saturated steam temperature at 70 bar is 285˚C which is greater than the reaction temperatures. Also, this allows us to stay consistent with the purchase of steam by only needing one steam pressure throughout the process. We used Equation 2 to solve for the flow rate of steam. The calculations were made based on the following information:

$\Delta h_{steam} = 1505\ kJ/kg$ (heat of vaporization of saturated 70 bar steam), an efficiency value of 0.80, and the heating duty for each unit (based on flowsheet from OLI). The following values were calculated:

R-100: $Q_{Process} = 410.48$ kJ/s

$$m_{steam} = \frac{410.48\ kJ/s}{1{,}505\ kJ/kg * 0.80} = 0.34 \text{ kg/s}$$

R-110: $Q_{Process} = 1{,}917.92$ kJ/s

$$m_{steam} = \frac{1{,}917.92\ kJ/s}{1{,}505\ kJ/kg * 0.80} = 1.59 \text{ kg/s}$$

R-130: $Q_{Process} = 64.38$ kJ/s

$$m_{steam} = \frac{64.38\ kJ/s}{1{,}505\ kJ/kg * 0.80} = 0.053 \text{ kg/s}$$

R-160: $Q_{Process} = 24.47$ kJ/s

$$m_{steam} = \frac{24.47\ kJ/s}{1{,}505\ kJ/kg * 0.80} = 0.02 \text{ kg/s}$$

R-170: $Q_{Process} = 16.53$ kJ/s

$$m_{steam} = \frac{16.53\ kJ/s}{1{,}505\ kJ/kg * 0.80} = 0.013 \text{ kg/s}$$

*Note heat duty was not accounted for in the calciner, as the OLI flow sheet would not converge when modeling the calciner unit. This will need to be examined in the future.

Appendix C: Capital Cost and Manufacturing Cost

Capital Cost Summary

Capital Cost Summary				
Equipment Number	Description	Capacity or Size Specs	Cbm	
Heat Exchangers				
E-101	Heat Exchanger: Cool R-100 Exit Stream	55.26 m^2, stainless steel, u-tube, 14.2 barg	74038.825	$
E-161	Heat Exchanger: Cool R-160 Exit Stream	4.51 m^2, stainless steel, u-tube, 10 barg	24621.4375	$
E-123	Heat Exchanger: Condense -122 Vapor Stream	18.49 m^2, stainless steel, u-tube, 10 barg	43423.2625	$
E-173	Heat Exchanger: Condense -172 Vapor Stream	18.27 m^2, stainless steel, u-tube, 10 barg	43423.26	$
Total heat exchangers			185506.785	$
Reactors				
R-100	Monazite-Acid Rxn	1.7 m ID x 6.8 m vert. stainless steel, jacketed vessel, 14.2 barg	964122.38	$
R-110	Cerium & Lanthanum Phosphate to Sulfate Rxn	1.9 m ID x 7.5 m vert. stainless steel, jacketed vessel, 14.2 barg	1075367.27	$
R-120	Neodymium Precipitation Rxn	5.4 m ID x 21.6 m vert. stainless steel, jacketed vessel, 34.4 barg	9136867.5	$
R-130	Neodymium Phosphate to Oxide Rxn	4.5 m ID x 17.9 m vert. stainless steel, jacketed vessel, 29.4 barg	7161082.5	$
R-170	Thorium Sulfate to Oxide Rxn	1.5 m ID x 5.9 m vert. stainless steel, jacketed vessel, 24.3 barg	896827.932	$
R-150	REE Calciner	Electrical circulation heater, 1.2kW	82457.15625	$
R-160	REE Sulfates to Hydroxide Rxn	1.5 m ID x 6.2 m vert. stainless steel, jacketed vessel, 5 barg	601021.05	$
R-1000	Dilution/Neutralization	1.7 m ID x 6.8 m vert. stainless steel, jacketed vessel, 5 barg	692480.775	$
R-180	Dilution/Neutralization	1.5 m ID x 5.9 m vert. stainless steel, jacketed vessel, 5 barg	517400.73	$
R-190	Waste Phase Conversion	0.73 m ID x 2.9 m vert. stainless steel, jacketed vessel, 5 barg	167304.375	$
R-140	Waste Phase Conversion	0.93 m ID x 3.7 m vert. stainless steel, jacketed vessel, 5 barg	227055.9375	$
Total Reactors			21521987.61	$
Process Vessels				
H-103	Splitter: Thorium Sulfate + Silicon Dioxide	Column: 0.82 m height, 0.21 m ID, stainless steel, Trays: 0.21m ID, stainless steel	19443.27225	$
H-132	Splitter: Solid Waste + Neodymium Hydroxide	Column: 1.02 m height, 0.26 m ID, stainless steel, Trays: 0.26 m ID, stainless steel	67809.94275	$
H-174	Splitter: Solid Waste + Thorium Oxide	Column: 0.54 m height, 0.14 m ID, stainless steel, Trays: 0.14 m ID, stainless steel	30740.908	$
H-163	Splitter: Solid Waste + Mixed REE Hydroxides	Column: 1.57 m height, 0.39 m ID, stainless steel, Trays: 0.39m ID, stainless steel	57375.1575	$
Total process vessels			175369.2805	$
Mixers				
M-100a	Agitator for R-100	2.64 kW, stainless steel	47421.875	$
M-110a	Agitator for R-110	3.43 kW, stainless steel	53112.5	$
M-120a	Agitator for R-120	42.68 kW, stainless steel	220037.5	$
M-130a	Agitator for R-130	27.29 kW, stainless steel	151750	$
M-170a	Agitator for R-170	1.93 kW, stainless steel	41731.25	$
M-160a	Agitator for R-160	2.13 kW, stainless steel	41731.25	$
M-1000a	Agitator for R-1000	2.64 kW, stainless steel	47421.875	$
M-180a	Agitator for R-180	1.91 kW, stainless steel	39834.375	$
M-190a	Agitator for R-190	0.36 kW, stainless steel	34143.75	$
M-140a	Agitator for R-140	0.63 kW, stainless steel	34143.75	$
Total mixers			711328.125	$
Pumps				
L-111		Centrifugal, 8.42 kW, stainless steel	56223.375	$
L-124		Centrifugal, 93.4 kW, stainless steel	172919.125	$
Total pumps			229142.5	$
Separators				
H-102	Filter: Thorium Oxide/Silicon Dioxide + REE Phosphates	Belt filter, 1.32 m^2, stainless steel	191205	$
H-162	Filter: REE Hydroxide + Liquid Waste	Belt filter, 1.33 m^2, stainless steel	191205	$
H-191	Filter: Solid Waste + Liquid Waste	Belt filter, 0.42 m^2, stainless steel	109260	$
H-141	Filter: Solid Waste + Liquid Waste	Belt filter, 1.30 m^2, stainless steel	126000	$
H-111	Settler: REE Phosphates + Neodymium Phosphate	Liquid cyclone, 5.4E-3 m^2, stainless steel	21245	$
H-121a		Liquid cyclone, 2.5E-4 m^2, stainless steel	4249	$
H-131a		Liquid cyclone, 3.4E-6 m^2, stainless steel	1396.1	$
H-122a		Liquid cyclone, 2.5E-4 m^2, stainless steel	4249	$
H-171a		Liquid cyclone, 6.9E-6 m^2, stainless steel	1396.1	$
H-172a		Liquid cyclone, 1.3E-4 m^2, stainless steel	3186.75	$
H-131b		Gas cyclone, 9.8E-5 m^2, stainless steel	8346.25	$
H-121b		Gas cyclone, 9.8E-5 m^2, stainless steel	8346.25	$
H-171b		Gas cyclone, 3.3E-5 m^2, stainless steel	8346.25	$
H-172b		Gas cyclone, 3.3E-5 m^2, stainless steel	8346.25	$
Total Separators			686776.95	$
Total bare module cost			23510111.25	$
Contingency and Fee			4231820.024	$
Total module cost			27741931.27	$
Auxiliary facilities			8322579.381	$
Grass-roots capital			**36064510.65**	$

Table C.1: Capital Cost Summary of the proposed process

Manufacturing Cost Summary

Job Title: Monazite Processing Plant

Capital	
Fixed Capital	$36,064,510.65
Working Capital	$3,606,451.07
Total Capital Investment	$39,670,961.72

Manufacturing Expenses	Annual Cost ($/yr)
Direct	
Raw Material	$13,859,667.99
Catalysts	$0
Operating Labor	$1,568,761.02
Supervisory Labor (15% of operating labor)	$235,314.15
Utilities	
Steam 63,747,386.96 kg/y 70 bar @ 0.05232$/kg	$3,335,144.62
Electricity 1,653,239.208 kWh @ 0.18791 $/kWh	$310,660.18
Refrigerant	$0
Cooling Water 4,397,534.61 m^3/y @ 0.184 $/m^3	$809,209.43
Wastewater Disposal	$3,861.95
Maintenance (6% of fixed capital)	$2,163,870.64
Operating supplies (15% of maint. & repairs)	$324,580.60
Laboratory charges (15% operating labor)	$235,314.15
Patents and Royalties (3% of total expense)	$685,391.54
Total	$23,531,776.27
Indirect	
Overhead (60% of of op labor, supervision, and maintenance)	$2,380,767.49
Local Taxes (2% of fixed capital)	$721,290.21
Insurance (1% of fixed capital)	$360,645.11
Total	$3,462,702.81
General Expenses	
Administrative Cost (25% of overhead)	$595,191.87
Distribution and Selling (10% of total expense)	$3,670,132.00
R&D (5% of total expense)	$1,835,066.00
Total	$6,100,389.87
Depreciation (10% of fixed capital)	$3,606,451.07
Total Expenses	$36,701,320.02
Revenue from Sales	$86,686,156.80
Net Annual Profit	$46,378,385.72
Income Taxes	$16,232,435.00
Net annual profit after taxes	$30,145,950.72

Table C.2: Manufacturing Cost Summary of the proposed process

Cash Flow Analysis Chart

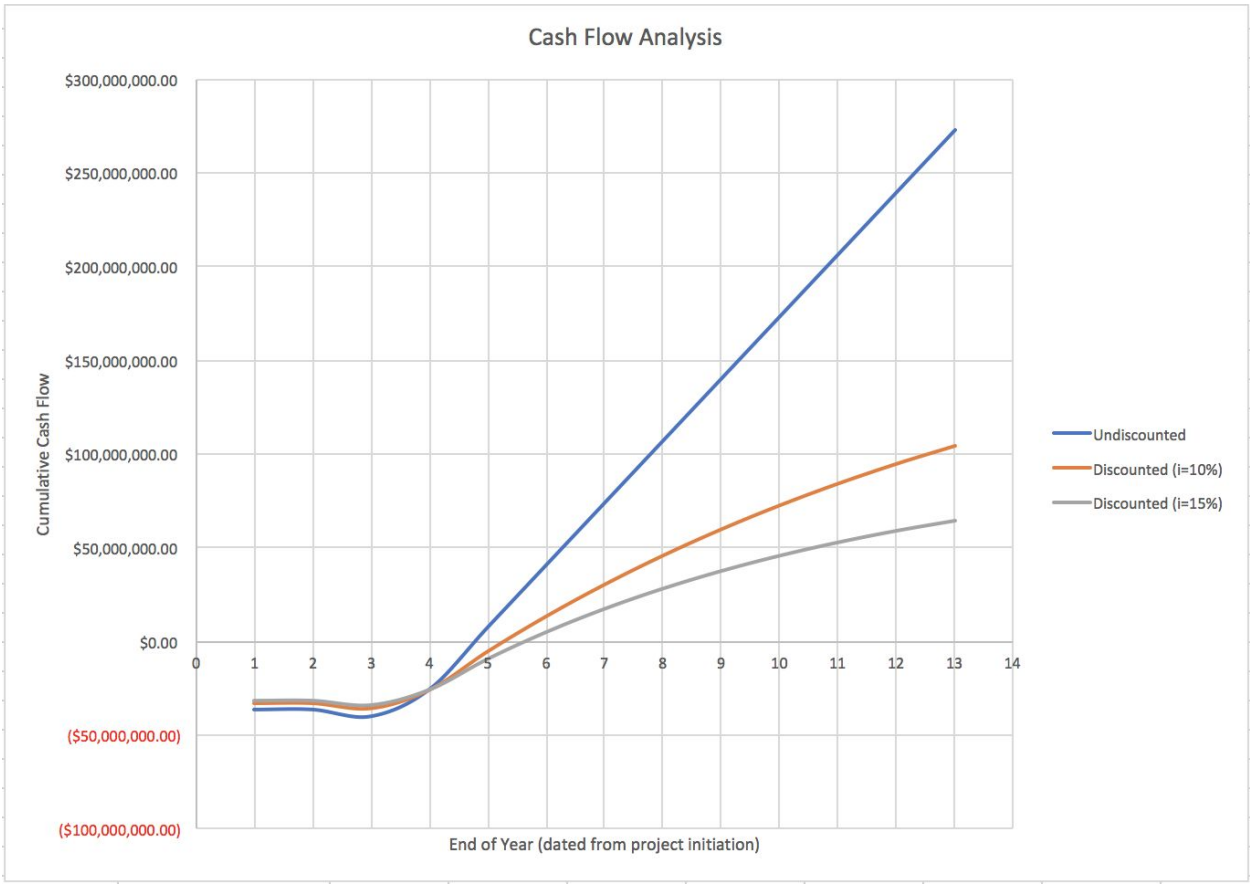

Figure C.1: Cash Flow Analysis Chart of the proposed process

Appendix D: Equipment Design and Cost

Reactor Calculation:

Sample- R-100 and M-100a

In order to get an accurate cost of the reactor, it must be designed properly. Based on the information obtained from Rodliyah, the reactor must have a space time of 2.5 hours to achieve optimal acid digestion. The process conditions of this reactor are 220˚C and 15 atm, based on the literature reactions and OLI calculations. We find the volumetric flow rate to the reactor consists of the raw material feed rate and the sulfuric acid feed rate. Once these parameters are known, volume and height can be determined for the reactor. The volume of the reactor will then allow us design and cost the agitator for the unit. We cost the vessel using figures 5.44, 5.45, and 5.46 from Ulrich. Agitator economic calculations incorporate figure 5.42 from Ulrich. It should be noted that stainless steel was used as the material of construction for the reactor and the agitator due to the fact that we are processing acid and the material is corrosion resistant. Also, it can handle high temperatures up to 920˚C (ssina.com). Ulrich notes a limitation for stainless steel processing acids at 200˚C, a deeper evaluation into this selection is suggested. We designed and costed the reactor by the following calculations:

$$V_{R-100} = q * \tau = 3.0339\ m^3/hr * 2.5hrs = 7.58m^3$$

$$V_{R-100}@\ 100\%\ excess = 7.58m^3 * 2 = 15.17m^3$$

$$D_{R-100} = \left(\frac{V_{R-100}*4}{2\pi}\right)^{1/3} = \left(\frac{7.58m^3*4}{2\pi}\right)^{1/3} = 1.69m$$

$$H_{R-100} = \frac{4*V_{R-100}@\ 100\%\ excess}{\pi D^2_{R-100}} = \frac{4*15.17m^3}{\pi*1.69^2 m} = 6.76\text{m}$$

$$P_{M-100a} = (V_{R-100} @\ 100\%\ excess)^{0.8} * 0.3 = (15.17 m^3)^{0.8} * 0.3 = 2.64 kW$$

*Note: Power equation comes from table 4.16 in Ulrich.

$$C_{bm,\,R-100} = C_P * F_{BM}(function\ of\ F_P\ and\ F_M) * 607/400 * jacket\ factor = \$26000 * 14.9 * 607/400 * 1.64 = \$964{,}122.38.$$

*Jacket factor was calculated by dividing jacketed vessels by plain vessels in figure 5.23 for volumes equivalent to the desired reactor volumes.

$$C_{bm,\,M-100A} = C_P * F_{BM} * 607/400 = \$12,500 * 2.5 * 607/400 = \$47,421.88$$

Other reactors and agitators

The other reactors implemented the same method. Parameters needs for design and costing can be obtained from OLI flowsheet. We assumed a space time of 2.5 hours for each vessel, and that the material of construction is stainless steel with extensive amount of acids and bases used in the process. Values are shown in the table below:

	Temperature (°C)	Pressure (atm)	q (m^3/hr)	Space Time (hrs)	Volume (m^3)	Volume (m^3, at 100% e	Diameter of reactor (m)	Height of reactor (m)	Cp (from book)index=400	Fm *stainless steel	Fp from book	Fbm *vertically oriented prc	Cbm, using 2018 chE (607)
R-100	220	15	3.033905	2.5	7.5847625	15.169525	1.690210141	6.760840565	26000	4	1.8	14.9	964122.38
R-110	275	15	4.19966	2.5	10.49915	20.9983	1.883698627	7.534794507	29000	4	1.8	14.9	1075367.27
R-120	250	35	68.26682	2.5	245.66705	491.3341	5.387769548	21.55107819	136000	4	2.9	22.3	9136867.5
R-130	250	30	56.176912	2.5	140.44228	280.88456	4.471562769	17.88625107	110000	4	2.75	21.45	7161082.5
R-170	220	25	2.05388	2.5	5.1347	10.2694	1.484106495	5.936425982	19800	4	2.3	18.2	896827.932
R-160	100	1	2.31685	2.5	5.792375	11.58475	1.564942301	6.179769204	23000	4	1.15	10.5	601021.05
R-1000	25	1	3.0346432	2.5	7.586608	15.173216	1.690347216	6.761388862	26500	4	1.15	10.5	692480.775
R-180	25	1	2.02209	2.5	5.055225	10.11045	1.476409636	5.905638544	19800	4	1.15	10.5	517400.73
R-190	25	1	0.247213	2.5	0.6180325	1.236065	0.7327634727	2.931053891	7000	4	1.15	10.5	167304.375
R-140	25	1	0.501002	2.5	1.252505	2.50501	0.9272991544	3.709196617	9500	4	1.15	10.5	227055.9375
R-150	360	1	0.036315	2.5	0.0907875	0.181575	0.3866357961	1.546543184	3460	4	1.15	10.5	82657.15625

	Volume of reactor (m^3)	power (kW)	Cp (mechanical seal)	fbm (ss)	cbm
M-100a	15.169525	2.641793336	12500	2.5	47421.875
M-110a	20.9983	3.426640658	14000	2.5	53112.5
M-120a	491.3341	42.67983973	58000	2.5	220037.5
M-130a	280.88456	27.2862308	40000	2.5	151750
M-170a	10.2694	1.933558476	11000	2.5	41731.25
M-160a	11.58475	2.129269276	11000	2.5	41731.25
M-1000a	15.173216	2.642307558	12500	2.5	47421.875
M-180a	10.11045	1.909579059	10500	2.5	39834.375
M-190a	1.236065	0.3554301805	9000	2.5	34143.75
M-140a	2.50501	0.6254157671	9000	2.5	34143.75

Table D.1: Values utilized in agitator design

Belt Filter Calculation:

Sample- Filter- H-102

Belt filters are used to separate solid-liquid mixtures at process temperatures less than 100˚C. In the case of H-102, the filter separated solid thorium oxide and silicon dioxide from liquid REE phosphates. The only parameter needed to determine the size and economics of the filter is the volumetric flow rate to the unit. The filter was designed using table 4.23b from Ulrich and the cost was determined using figure 5.57b from Ulrich. The belt filter was designed to be constructed of stainless steel since corrosive acids will be present in the unit. It was also assumed that this is a continuous filter. The design and cost calculations are shown below:

$$A_{Nominal\ area\ of\ H-102} = \frac{q}{design\ factor} = \frac{6.6*10^{-4} m^3/s}{5*10^{-4} m/s} = 1.32 m^2$$

*Note: Design factor is based on feed liquid flow rate from table 4.23b in Ulrich. The material was assumed to resemble dilute slimes.

$$C_{bm,\,H-102} = C_P * F_{BM} * 607/400 = \$35000 * 3.6 * 607/400 = \$191,205$$

Other belt filters

H-162 was designed using the same method and assumptions as H-102. H-141 and H-191 were designed using the same method, however one assumption was different. Instead of resembling dilute slimes, these filter process material that more closely resembles sands in table 4.23b in Ulrich.

Flotation Column Calculation:

Sample- H-103

Floatation columns were used to separate solids in the process by their densities. In this case, we are separating silicon dioxide from thorium sulfate. After consulting Dr. Counce, we decided to design the flotation column as a vertically oriented process vessel with two sieve trays. We designed the vessel and its tray to be stainless steel to be safe because thorium is in sulfate form and the process unit upstream contains a sulphuric acid digestion. The parameters needed in the calculation are volumetric flow rate to the column, process conditions, and spacetime.We cost the vessel using figures 5.44, 5.45, and 5.46 from Ulrich. We cost the sieve trays by figure 5.48 from Ulrich. The design and cost calculations are provided below:

$$V_{H-103} = q * \tau = 0.02743\ m^3/hr * 0.5hrs = 0.013715\ m^3$$

$$V_{H-103}@\ 100\%\ excess = 0.013715\ m^3 * 2 = 0.02743\ m^3$$

$$D_{H-103} = \left(\frac{V_{H-103}*4}{2\pi}\right)^{1/3} = \left(\frac{0.013715m^3*4}{2\pi}\right)^{1/3} = 0.2059m$$

$$H_{H-103} = \frac{4*V_{H-103}@\ 100\%\ excess}{\pi D^2_{H-103}} = \frac{4*0.02743m^3}{\pi*0.2059^2 m} = 0.824\ \text{m}$$

$$C_{bm,\,H-103,\,vessel} = C_P * F_{BM}(function\ of\ F_P\ and\ F_M) * 607/400 = \$1175 * 10.5 * 607/400 = \$18722.156$$

$$C_{bm,\,H-103,\,trays} = C_P * F_{BM} * N_{act} * f_q * 607/400 = \$30 * 2.2 * 2 * 3.6 * 607/400 = \$721.12$$

$$C_{bm,\,H-103,\,total} = \$19,443.28$$

Other flotation columns

The same method is used to calculate the other three flotation columns. Parameters of interest come from OLI flowsheet data.

Heat Exchanger Calculation:

Sample- E-173

This unit has a process stream that is predominantly water vapor at 160˚C at 1 atm, with 1.17 wt% of ammonia and is coming out of the exchanger as a liquid at 25˚C. An assumption that the process stream is all water was made to simplify calculations. The heat duty, mass flow rate of process stream, inlet and outlet temperature of the process stream, and assumed inlet and outlet temperature of cooling water stream are the parameters needed for the design. Since the process stream could contain a moderately strong base, stainless steel is used for the material of construction. We calculated the surface area needed for a U-tube heat exchanger and used figures 5.36, 5.37, and 5.38 for Ulrich to find economics for the equipment. It is presumed that there will be three sections in the heat exchanger; two subcooling sections and a condensation section. The calculation is demonstrated below:

First subcooling section: Process stream (160 ˚C vapor to 100˚C vapor) assuming cooling water comes in at 9˚C and leaves at 24˚C.

$$Q = (h_{at\ 160^{\circ}C,vap} - h_{at\ 100^{\circ}C,vap}) * m_{process\ stream} = (2,795.8 - 2,675.9)kJ/kg * 0.557\ kg/s = 66.78kJ/s$$

*Enthalpy values obtained from steam tables

$$\Delta T_m = \frac{(160-24)-(100-9)}{ln((160-24)/(100-9))} = 112^{\circ}C$$

$$A_{sub1} = \frac{Q}{U*\Delta T_m} = \frac{66.78\ kJ/s*1{,}000J/KJ}{900\ J/m^2s^{\circ}C * 112^{\circ}C} = 0.66\ m^2$$

*The U value was obtained from table 4.15a in Ulrich.

Condensation section: Process stream (100 ˚C vapor to 100˚C liquid) assuming cooling water enters at 9˚C and leaves at 24˚C.

$$Q = (h_{at\ 100^{\circ}C,\ latent\ heat\ of\ vaporization}) * m_{process\ stream} = 2,257.6\ kJ/kg * 0.557\ kg/s = 1,257.48\ kJ/s$$

*Enthalpy value obtained from steam table

$$\Delta T_m = \frac{(100-24)-(100-9)}{ln((100-24)/(100-9))} = 83.3^{\circ}C$$

$$A_{cond} = \frac{Q}{U*\Delta T_m} = \frac{1,257.48\ kJ/s*1,000J/KJ}{1,200\ J/m^2s^{\circ}C * 83.3^{\circ}C} = 12.58\ m^2$$

*The U value was obtained from table 4.15a in Ulrich.

Second subcooling section: Process stream (100 ˚C liquid to 25˚C liquid) assuming cooling water comes in at 9˚C and leaves at 24˚C.

$$Q = (h_{at\ 100^{\circ}C,liq} - h_{at\ 25^{\circ}C,liq}) * m_{process\ stream} = (417.51 - 104.84)kJ/kg * 0.557\ kg/s = 174.16\ kJ/s$$

*Enthalpy value obtained from steam table

$$\Delta T_m = \frac{(100-24)-(25-9)}{ln((100-24)/(25-9))} = 38.51^{\circ}C$$

$$A_{sub2} = \frac{Q}{U*\Delta T_m} = \frac{174.16\ kJ/s*1,000J/KJ}{900\ J/m^2s^{\circ}C * 38.51^{\circ}C} = 5.02\ m^2$$

*The U value was obtained from table 4.15a in Ulrich.

Total area for heat exchanger = $A_{sub1} + A_{cond} + A_{sub2} = 18.27\ m^2$

$$C_{bm,\ E-173} = C_P * F_{BM}(function\ of\ F_P\ and\ F_M) * 607/400 = \$4,850 * 5.9 * 607/400 = \$43,423.26$$

Other heat exchangers

The U-tube, E-123, is calculated in the same fashion as E-173. The process stream's main component is water vapor, only this time it is coming into the exchanger at 190˚C and leaves as a liquid at 25˚C. The same assumptions are made to obtain the design and cost.

E-101 and E-161 only have one subcooling section in the heat exchanger. These exchangers use the heat duty from OLI as the value of Q in the problem. E-101 assumes cooling water coming in at 9˚C and leaving at 23˚C. While E-161 assumes cooling water coming in at 9˚C and leaving at 16.16˚C. OLI can provide needed process stream conditions and heat duty.

Cyclone Calculation:

Separators modeled in OLI with three outlet streams represent one gas cyclone unit and one liquid cyclone unit in series. The series orientation allows the solid, liquid, and vapor phases to be separated in individual outlet streams. These separators are units H-121, H-131, H-171, and H-172.

Sample- Liquid Cyclone- H-121a

Cyclones are low maintenance, inexpensive units that require little to no operator attention. The liquid cyclone separate solids from liquids and can handle temperatures up to 400˚C. The preliminary design and economics of the cyclone are like belt filters in that they only depend on the volumetric flow into the unit. Stainless steel was used as the material of construction due to the presence of corrosive sulfates in the process stream. A more in-depth design of these types of separators is needed before moving further in the project. The preliminary design is based on table 4.23a in Ulrich and the economics of the liquid cyclone is based on reading figure 5.55 in Ulrich and using the required volumetric flow rate. The sample calculation is provided below:

$$A_{Nominal\ area\ of\ H-121a} = \frac{q}{design\ factor} = \frac{1.247*10^{-3} m^3/s}{5 m/s} = 2.49 * 10^{-4}\ m^2$$

*Design factor is based on feed liquid flow rate from table 4.23a in Ulrich.

$$C_{bm,\,H-121a} = C_P * F_{BM} * 607/400 = \$700 * 4 * 607/400 = \$4,249.00$$

Sample- Gas Cyclone- H-171b

Gas cyclones in our process separate liquids from gases. Like belt filters and liquid cyclones, the design and economics only depend on the volumetric flow rate. Stainless steel was again used due to the presence of corrosive sulfates. The preliminary design is based on table 4.23a in Ulrich and the economics of the liquid cyclone is based on reading figure 5.56 in Ulrich and using the required volumetric gas rate. The sample calculation is provided below:

$$A_{Nominal\ area\ of\ H-171b} = \frac{q}{design\ factor} = \frac{5.015*10^{-4} m^3/s}{15 m/s} = 3.34 * 10^{-5}\ m^2$$

*Design factor is based on feed liquid flow rate from table 4.23a in Ulrich.

$$C_{bm,\,H-171b} = C_P * F_{BM} * 607/400 = \$1,100 * 5 * 607/400 = \$8,346.25$$

Other cyclones

The other cyclones use the same assumptions and methods as what the sample examples implemented.

Pump Calculation:

Sample- L-124

In order for flow to subsist, sometimes energy must be added to the liquid to increase the pressure at that reference position above downstream process pressure. In this case we examine stream 3B. The design of a pump requires knowledge of the desired volumetric flow rate, desired inlet and outlet pressures of the pump, viscosity of the material, and motor efficiency. OLI

provides us with the flow rates and desired pressures. Further research is needed for the remaining parameters. The economics for the pump were determined from figures 5.49, 5.50, and 5.51 from Ulrich. We chose a centrifugal pump because of its ability to handle solids (as we have some in the process stream). Stainless steel was used as the material of construction because of its corrosion resistance. The pump has desired inlet and outlet pressures of 15 atm and 36 atm respectively. The calculation used for the pump design and its economics is provided below:

$$\varepsilon\,(pump\ efficiency) = (1 - 0.12q^{-0.27})(1 - \mu^{0.8}) = (1 - 0.12(0.0271 m^3/s)^{-0.27})(1 - (0.0014\ Pa\ s)^{0.8} = 0.6786$$

*Stream viscosity obtained from sulphuric-acid.com for a 44 wt% acid by weight stream.

$$W_s = \frac{q\Delta P}{\varepsilon} = \frac{0.0271\ m^3/s * (3{,}647{,}700\ Pa - 1{,}519{,}875\ Pa)}{0.6786} = 84,980.83\ Watts$$

$$Motor\ power\ consumption = \frac{W_s}{\varepsilon_{\ motor}} = \frac{84{,}980.83\ Watts}{0.91} = 93,385.52\ Watts$$

* ε for the electric motor was estimated from figure 4.2 in Ulrich.

$$C_{bm,\,L-124} = C_P * F_{BM}(function\ of\ F_P\ and\ F_M) * 607/400 = \$21,500 * 5.3 * 607/400 = \$172,919.13$$

Centrifugal Pump-111

Design was carried out using the same method and construction material. Flow rates and stream composition differs from 3b.